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CONTRACT DSS NUMBER 23440-7-9212

MINIMIZATION/UTILIZATION OF MINERAL INDUSTRY EFFLUENT TREATMENT SLUDGES

GML PROJECT No 7467

FOR

CANADA CENTRE FOR MINERAL AND
ENERGY TECHNOLOGY
ENERGY, MINES AND RESOURCES
CANADA

DSS CONTRACT No 15SQ.23440-7-9212

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JUNE, 1989

59 pp
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SUMMARY

The Canadian Mineral Industry generates large volumes of effluent treatment sludges, some of which may be classified as hazardous wastes based on previous studies by GML. This study considered a number of means for dealing with the problem, including sludge dewatering, treatment of effluents by ion exchange with subsequent metal recovery, and reprocessing of sludges either on-site by the use of solvent extraction, or in mineral industry smelters and refineries. Flowsheets for these processes were derived, and cost estimates prepared, which indicate the following:

- The alternatives to conventional lime treatment will all result in significant cost increases to the mineral industry..
- Both capital and operating costs are greatly affected by contaminant concentrations.
- The costs for the ion exchange and solvent extraction schemes considered may be lower in some cases than combined conventional lime precipitation, sludge dewatering, and disposal in a maximum secure landfill.

The following technology development needs are recommended:

- Develop more efficient processes to recover heavy metals directly from effluents. This may include electrochemical, ion exchange, and biological processes.
- Develop and implement methods to minimize metal release at source.
- Evaluate and implement means to maximize the recycle of sludges to Mineral Industry smelters and refineries.

SOMMAIRE

L'industrie canadienne des mines génère de gros volumes de boues provenant du traitement des effluents dont certaines sont catégorisées comme déchets dangereux selon des études précédentes de GML. La présente étude a évalué un certain nombre de moyens pour faire face à ce problème, incluant l'asséchage des boues, le traitement des effluents par échange d'ions avec récupération subséquente des métaux, et le recyclage des boues, soit sur le site même par extraction avec des solvants ou dans des industries minières comme les fonderies et les raffineries. Les diagrammes de principe qui ont été développés pour ces procédés de même que les estimés de coûts qui s'y rattachent, indiquent les points suivants:

- Les solutions alternatives au traitement conventionnel à la chaux ont toutes comme conséquence une augmentation significative des coûts pour l'industrie minière.

- Les coûts de capital et d'opération sont en relation directe avec les concentrations en contaminants.

- Les coûts pour les méthodes évaluées d'échange d'ions et d'extraction avec des solvants peuvent être inférieurs, dans certains cas aux coûts combinés de la précipitation à la chaux suivie d'un asséchage des boues et d'une disposition finale dans un site à sécurité maximale.

Le développement des technologies suivantes est recommandé:

- Développer des procédés plus efficaces pour récupérer les métaux lourds directement des effluents. Ceci pourrait inclure des procédés électrochimiques, à échange d'ions et biologiques.

- Développer et implanter des méthodes afin de minimiser le dégagement des métaux à la source.

- Evaluer et implanter des moyens pour maximiser le recyclage des boues dans des industries minières comme les fonderies et les raffineries.

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1.0 INTRODUCTION

The Canadian Mineral Industry has in use numerous effluent treatment facilities which generate large volumes of waste sludges. A previous study by Gionet, Mellor, Liebich Associates Ltd. (GML) has shown that 50% of sludge samples from a cross section of Canadian base and precious metal producers can be classified as hazardous wastes based on Québec regulatory criteria. Recent hazardous waste regulations being implemented by Canadian provincial regulatory agencies may result in high costs for environmentally acceptable disposal of these wastes. Current disposal methods may be inadequate and sludges already disposed may require further treatment. Additionally, metal values in some of the sludges may allow economic metal recovery.

In order to evaluate available options for dealing with the problem of mineral industry effluent treatment sludges, GML has undertaken a project sponsored and funded by the Mineral Sciences Laboratories of CANMET. The study deals mainly with base metal producers, which generate the majority of sludges. The main aims of the study are to define sludge disposal options, and indicate the research needs for:

- Minimization of sludge quantities by the replacement or modification of the lime precipitation process currently in use.

- Internal recycle of sludges to milling processes.

- External recycle of sludges as smelter/refinery feeds.

2.0 BACKGROUND

The Canadian Mineral Industry uses lime to neutralize acidity, and precipitate heavy metals from base metal mine, smelter, and refinery effluents[1]. In the milling of precious metals, cyanide destruction using a number of oxidation methods is practised[2]. The results of the previous GML study[3] have shown that the sludges produced during such treatment contain a variety of contaminants. Calcium, iron, and zinc were the three main elements present in most of the sludges. Copper, lead, nickel, aluminium, magnesium, manganese, arsenic, cadmium, sodium and antimony were less common, and were generally found in lower concentrations.

Leaching studies were conducted using both regulatory and GML protocols [3] on the sludges with a variety of leaching media (distilled water, sulphuric acid, acetic acid/sodium acetate). The results of the leaching studies (as presented in Appendix 1) indicated the following:

- The release of contaminants from the sludges was related to the final pH of the leachates. The use of acetic acid/sodium acetate solution did not result in higher contaminant concentrations than sulphuric acid.

- Contaminant release in leach liquor was also related to initial concentration of metals in the sludge. Thus, lead, cadmium, nickel, and arsenic were present in low concentration in most sludge samples, as well as leach solutions. Where these elements were present in sludges at levels above 0.5%, the low pH leach solutions contained significantly higher contaminant concentrations.

- The highest release and leach solution concentrations were for zinc, iron and copper, with levels as high as 5200 mg/L Zn, 770 mg/L Fe, and 93 mg/L Cu.

- Leaching of old and fresh sludges appeared to result in similar leach contaminant concentrations.

Most regulatory requirements define a hazardous waste using leaching tests as a criterion. The Quebec Hazardous Waste Regulations[4] and Ontario Regulation 309[5] define a hazardous waste as one whose leachate concentrations exceed the quantities indicated in Table 1. A comparison of these values with the leaching test results has indicated Zn, Ni, Cd, Cu to be elements exceeding the Quebec regulatory requirements in up to 50% of the sludge samples tested.

TABLE 1

WASTE LEACHATE PARAMETERS RELATED TO MINERAL INDUSTRY SLUDGES

PARAMETER	WASTE HAZARDOUS IF FOLLOWING CONCENTRATION (mg/L) EXCEEDED	
	Quebec[4]	Ontario[5]
Arsenic	5.0	5.0
Barium	---	100.0
Boron	---	500.0
Cadmium	2.0	0.5
Chromium	5.0	5.0
Copper	10.0	---
Cyanide (Free)	---	20.0
Nickel	10.0	5.0
Lead	5.0	5.0
Zinc	10.0	---

3.0 MINIMIZATION OF SLUDGE QUANTITIES

Large sludge volumes are currently produced by mineral industry effluent treatment systems due to the thixotropic nature of metal hydroxide sludges, which may contain up to 98% water content. This section considers possible means to reduce sludges quantities.

3.1 METAL PRECIPITATION REAGENTS

3.1.1 Hydroxide Precipitation

Current wastewater treatment by Canadian base and precious metal producers involves mainly pH adjustment and removal of heavy metals (eg. copper, lead, nickel and zinc). In some effluents, arsenic, antimony and cyanides are also removed. The most common method of treatment due to its low cost, is the addition of lime, either in the hydrated or unslaked form. Heavy metals, present mostly as sulphates are typically precipitated as follows:

$$2H_2O + MSO_4 + Ca(OH)_2 \longrightarrow M(OH)_2 + CaSO_4 \cdot 2H_2O$$

where M represents metals (eg. Cu, Zn, Fe). Precipitated solids consist mainly of metal hydroxides, gypsum, and unreacted lime. Clarifier underflow may range from 3 to 20% solids concentration.

NaOH is used by many metal finishing operators in effluent treatment due to its ease of handling (direct addition of liquid NaOH instead of need to handle dry lime, and prepare lime slurries). Lower sludge volume production may occur with

NaOH treatment of effluents containing high sulphate concentrations due to the higher solubility of $Na_2 SO_4.10 H_2O$ vs that of $CaSO_4.2H_2O$. Sulphates are generally not precipitated when NaOH is used in effluent treatment, resulting in sulphate and sodium discharge in the final effluent. The hydroxide sludge produced by NaOH tends to be more thixotropic, however, with metal hydroxide sludges of 2 to 3% solids concentrations generally achieved in clarifier underflow. Both lime and sodium hydroxide generally provide effluents with metal residuals in the 0.5 to 1 mg/L range. Stricter regulatory requirements may make sulphide precipitation more attractive.

3.1.2 Sulphide Precipitation

The precipitation of heavy metals by sulphides offers some advantages. Metal sulphides are less soluble than hydroxides, resulting in lower dissolved metals remaining in solution. Hydroxides tend to have a minimum solubility which varies with each metal by pH. The solubility of sulphides declines uniformly with pH, however as shown in Figure 1. Thus, a single pH set-point can result in efficient precipitation of a variety of metals.

Use of sulphide precipitation for the treatment of metal containing effluents has been evaluated in a number of previous studies[6-10]. Heavy metals, as sulphates would be precipitated as follows using sodium sulphide:

$$MSO_4 + Na_2S \longrightarrow MS + Na_2SO_4$$

where M represents metals (eg Cu, Zn, Fe). The method has a number of disadvantages. While pH control can be used to control lime or caustic addition, no easy method has been shown for the automatic doseage control of sulphide reagents.

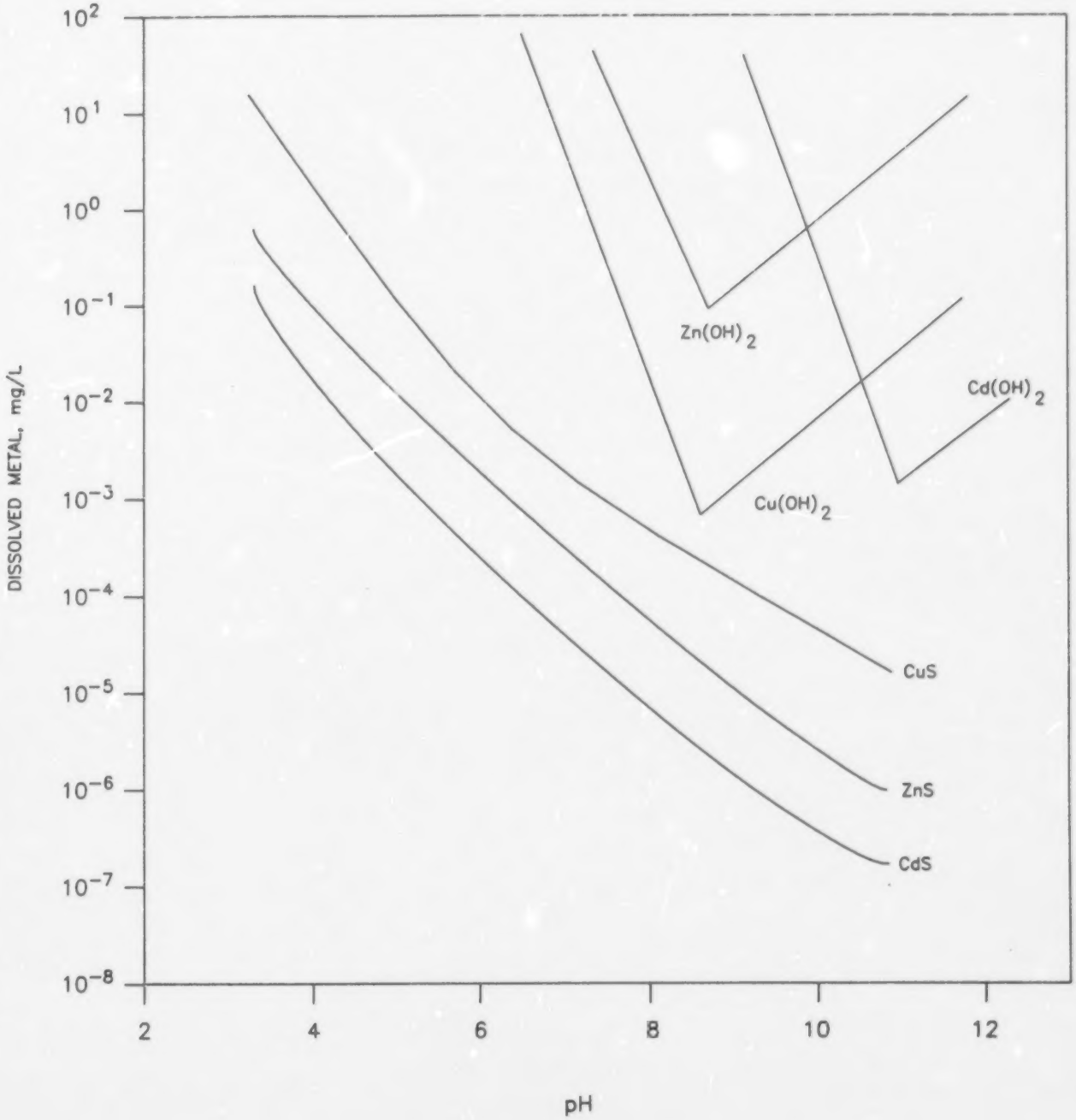

FIGURE 1: METAL SULPHIDE AND HYDROXIDE SOLUBILITY CURVES

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Also excess sulphide presents the hazard of hydrogen sulphide formation as follows:

$$1)\ Na_2S + H_2O \longrightarrow 2\ Na^+ + S^{-2} + H_2O$$

$$2)\ 2\ Na^+\ S^{-2} + 2H_2O \longrightarrow 2OH^- + H_2S\uparrow + 2\ Na^+$$

Thus, closed reactor vessels with secure venting would be required to minimize safety risks. Little information exists on sludge settleability or dewatering characteristics of sulphide sludges, although there is indication of poor settling.

3.2 SLUDGE DEWATERING

GML's previous study[3] has found that mineral industry effluent treatment plants with conventional solid/liquid separation (either mechanical clarifiers or settling ponds), achieve sludge solids concentrations up to 5%. The High Density Sludge[11] mode is in use at a number of locations. As shown in Figure 2, the main feature of this method is the partial recycle of sludge to mix with lime slurry prior to flow to the neutralization reators. At locations where sludge recycle is practised, sludge solids concentrations of 15 to 20% are attained.

Siwik[12] reviewed examples of land disposal of high density sludges at Noranda's Waite Amulet and Mines Gallen, and Cominco's Sullivan operations. At both Cominco, and Mines Gallen sites, a sand and gravel bed is used, while an engineered drainage system consisting of sand and gravel filters and drainage tiles is installed at Waite Amulet. Kuit[13] has previously reported the Sullivan site to achieve 50% solids upon dewatering.

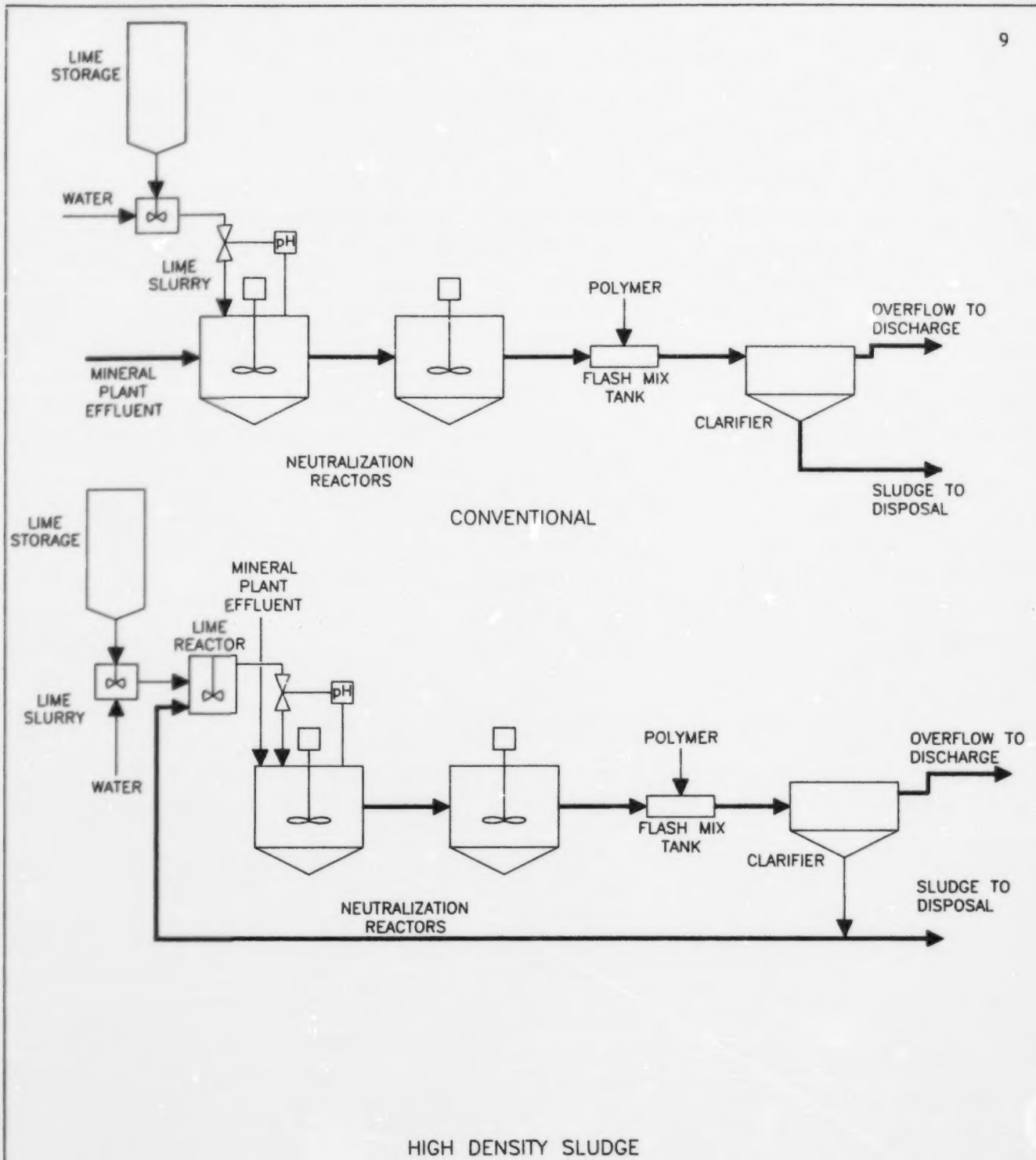

FIGURE 2: CONVENTIONAL AND HIGH DENSITY SLUDGE FLOWSHEETS

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The dewatering of metal hydroxide sludges using mechanical filters is coming into wide use in the surface finishing and plating industries faced with high waste disposal costs. In the Canadian Mineral Industry, two typical examples are:

- Cominco's Trail facility[14] operates a 2.4 m diameter, 3.6 m long drum vacuum filter to dewater underflow solids from a clarifier.

- Inco's Port Colbourne refinery[15] dewaters effluent treatment sludges prior to disposal to a land fill. A drum vacuum filter in use produces 45% solids filter cake.

Previous work by Crocker[16] and Campbell[17] has indicated filter cake of 35 to 45% solids achievable using rotary drum vacuum filters for metal hydroxide sludge dewatering. Filter yields of 10 to 25 $kg/m^2/hr$ are indicated.

Dryers can be used to dewater metal hydroxide sludges to below 10% moisture. Direct heat units are the simplest and most economical in construction, and are used where direct contact between solids and hot air or gases can be tolerated. Because the total heat load must be introduced in the gas stream, large gas volumes and high gas velocities are usually employed causing entrainement of dust from the dried product. Thus, air cleaning equipment (cyclones, filters, or scrubbers) is usually required.

Indirect heat units utilize a secondary heat transfer agent (eg. steam or thermal fluid) which does not come in contact with the material dried. Screw type dryers can be completely sealed to eliminate the need for any auxillary dust collection equipment.

These units dry the material by transferring heat from the heat transfer agent through the surfaces of the hollow flights of a screw and dryer jacket. The continuous movement of the screws conveys and dispenses the material through the troughs. This causes the particles to come into frequent contact with the surfaces containing the heat transfer agent, resulting in efficient heat transfer.

3.3 SLUDGE DEWATERING DESIGN CONSIDERATIONS

Assuming that the High Density Sludge mode of operation is in use, a plant using lime can be expected to achieve 20% solids in the clarifier underflow. There is no data to show that this mode of operation can increase the concentration of sodium hydroxide, or sodium sulphide sludges. Experiments by GML in a metal finishing effluent treatment plant removing zinc and iron with sodium hydroxide have shown no increase in sludge concentration when sludge recycle was attempted.

Industry data suggests that use of vacuum filters for dewatering of lime sludges will achieve a 45% to 50% filter cake. Sodium hydroxide sludges generally dewater to only 20% in vacuum or pressure filters. Little information is available on dewatering of sodium sulphide sludges.

4.0 METAL RECOVERY PROCESSES

4.1 ION EXCHANGE PROCESS

4.1.1 Background

Ion exchange generally involves the use of synthetic organic resins in columns. A solution containing impurities is contacted with the resin, usually by downward flow through the resin bed. As the solution contacts the resin, the ions to be removed exchange with mobile ions in the resin. The process continues until an equilibrium or breakthrough point is reached, when excessive concentration of impurities leave the column. The resin can then be regenerated to its original state by the use of acid, base, or salt solutions. The regenerated resin is then ready for additional service.

The ion exchange process has been used in the mineral industry principally to concentrate metals in solutions by extraction at low concentration levels, and stripping (regenerating) at a high concentration. Uranium is commercially extracted from leach liquors by anion exchange, then concentrated as well as purified by elution with a solution such as HNO_3 - NH_4NO_3 .

An ion exchange process has been evaluated for the removal of molybdenum from Climax mine waste waters [18], with pilot scale tests conducted using a continuous countercurrent ion exchange unit. The molybdenum removed by the ion exchange unit was to be processed by solvent extraction-crystallization to produce marketable sodium molybdate.

Environment Canada have sponsored an evaluation of the ion exchange process for metal recovery at several metal finishing operations. Generally, the schemes evaluated provide recovery of only one metal at a time from plating or rinse solutions.

A chromic acid ion exchange purification system supplied by Eco-Tec Ltd. was installed at a hard chromium electroplating operation in Winnipeg, Man. (Canadian Bronze Company Ltd)[19]. The unit was used for the removal of iron build-up in chromic acid plating baths, allowing these solutions to be re-used. A flow of 30 L/hr was treated, reducing iron from 11.9 to 8.3 g/L. A bleed stream containing iron and chrome was sent for conventional chemical treatment (hexavalent to trivalent chrome reduction, followed by precipitation of metals as hydroxides).

A job plating shop in Montreal (G.G. Buffing and Plating) installed systems by Eco-Tec Ltd. for separate copper, nickel, and chrome rinses[20]. The process consists of passing rinse water through the ion-exchange units to concentrate the metals for recycle to the plating baths. Typical values achieved are as follows:

	Chrome Line Cr (mg/L)	Nickel Line Ni (mg/L)	Copper Line Cu (mg/L)
Bath	260,000	84,000	82,000
Rinse	170	460	260
Recovered Product	53,000	22,000	35,000

Eco-Tec Ltd. have installed a system at a brass milling operations, to treat and recover a combined stream of copper and zinc from rinse waters and pickling bath purification[21]. The solutions are filtered, then passed through an ion exchange bed to remove copper and zinc. The resulting effluents contain 1 mg/L Cu and 1 mg/L Zn. The ion exchange

bed is regenerated using sulphuric acid, which results in a concentrated stream consisting of copper sulphate, zinc sulphate, and dilute sulphuric acid. This solution is sent for copper electrowinning. The spent electrolyte is pH adjusted with NaOH, and passed through a second ion exchange bed to remove any residual copper from the solution. The resulting copper free zinc salt solution is then sent for zinc electrowinning, and the spent electrolyte is recycled.

Dianex Systems Ltd. of Brantford, Ont. market the Diaion series of ion exchange resins manufactured by Mistubishi Chemical Industries Ltd. Some of its resins are aimed specifically at heavy metal removal and recovery from waste waters. The company conducted laboratory scale tests for a U.S. mining company for the removal of Cu, Pb, Zn, using the Diaion resins[22]. The mine treats its effluents with lime, and is unable to achieve regulatory levels. The ion exchange system is considered for polishing treatment following lime addition and sedimentation. Up to May 1989, no work beyond bench scale tests was carried out.

4.1.2 Ion Exchange Design Considerations

Ion exchange processes have been successful in the removal of copper, zinc, cadmium, nickel and cobalt to less than 1 mg/L from effluent streams. The common (relatively cheap) resins used in water softening etc., are not ion selective, and high concentrations of calcium and iron in relation to heavy metals would result in rapid resin depletion. Chelating resins recommended for these applications are less selective for iron (with removals of 20%), and do not absorb sodium, calcium, magnesium, or sulphate. These ion selective resins have much lower inherent life spans however than conventional resins, and are far more expensive ($3500/$m^3$ of non selective resin vs $35,000/$m^3$ of selective resins).

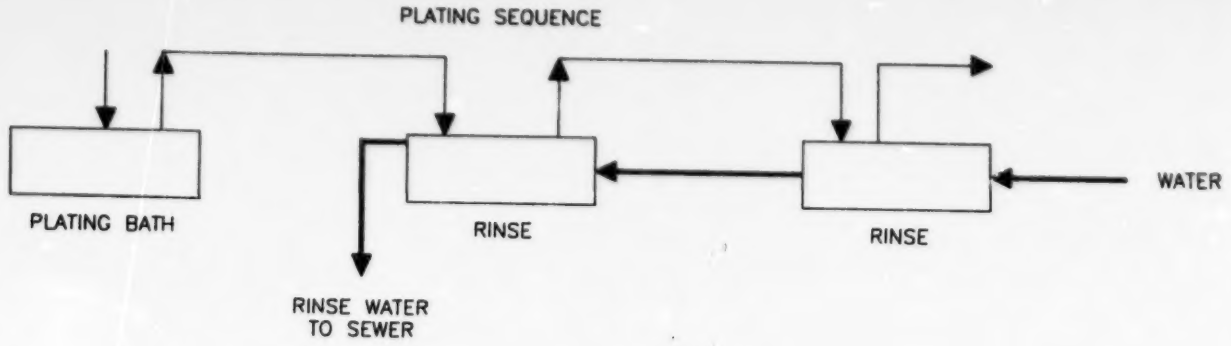
When the resin becomes exhausted, the absorbed cations are eluted with either sulphuric or hydrochloric acid. In some cases, after elution it may be necessary to reconvert to the free base form by passing a solution of sodium hydroxide through the column. The cation containing elute may be processed through other ion exchange columns for more selective metal removal, or through electroplating or precipitation techniques for the recovery of metals.

Typically, metal loading rates of 24 to 48 kg/m^3 are achieved. Elution with sulphuric acid requires 120 kg H_2SO_4/m^2 resin (generally, a 10% acid solution is used) followed by 9 to 10 bed volumes of demineralized rinse water. Hydraulic loadings required in ion exchange beds are 2.3 $m^3/hr/m^2$. Bed depth depends on the concentration of inlet contaminants, and expected discharge.

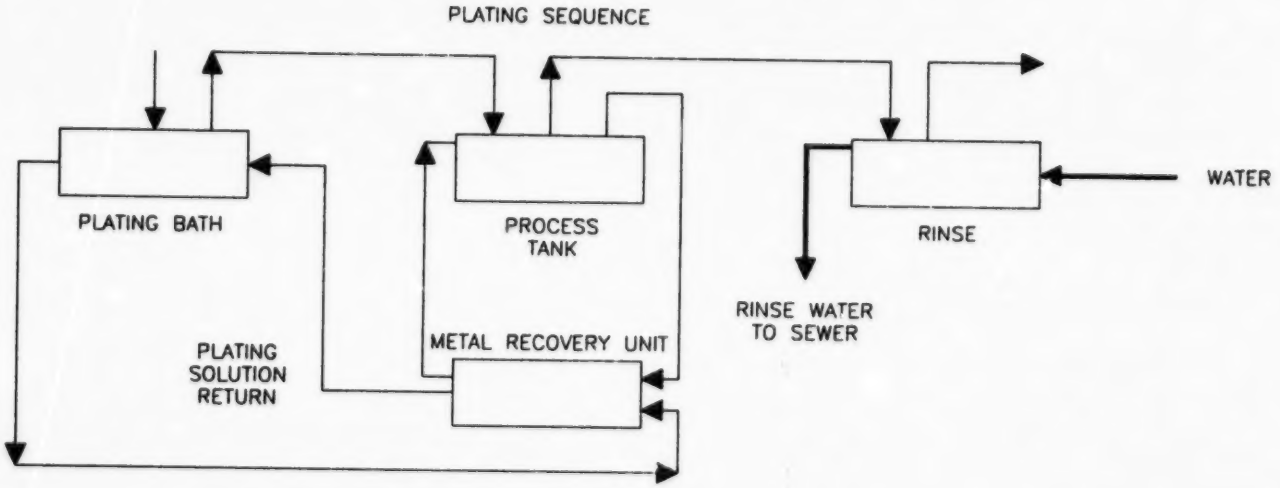
4.2 ELECTROCHEMICAL PROCESSES

Electrochemical processes are widely used by the Canadian Mineral Industry for the refining of base and precious metals. To date, there has been little use of electrochemical processes in effluent treatment. Most effluent streams contain relatively low metal concentrations which make direct application of the technology costly both in capital and operating costs.

Electrochemical systems have been used in metal finishing operations for the recovery of metals from plating solutions. Typical schemes showing operation with and without metal recovery are presented in Figure 3.



PLATING SEQUENCE WITHOUT METAL RECOVERY



PLATING SEQUENCE WITH METAL RECOVERY

FIGURE 3: METAL RECOVERY IN PLATING OPERATIONS

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Environment Canada has sponsored a number of technical evaluations of electrochemical technologies in metal finishing as follows:

The HSA reactor technology utilizes a carbon fiber cathode, and a metallic anode. The technology is claimed to result in vastly superior performance over other reactor types. Several of the systems manufactured by HSA Reactors Ltd. have been installed since 1981 with varying degree of success. Six installations of the HSA system were tested under a recent joint EPA/Environment Canada study[23]. The results showed a wide variation in performance, ranging from nearly ineffective, to a metal removal and cyanide destruction capability sufficient to achieve compliance with effluent regulations. Zinc and cadmium plated during operations were found to contain 98.7 to 99.4% Zn, and 99.3 to 99.9 % Cd making these highly recycleable to primary and secondary metal markets.

The Chemelec system supplied by BEWT Recovery Technologies Inc. consists of an electrolysis chamber fitted with mesh electrodes sited within a bed of inert glass beads which are fluidized by the action of the pumped liquid. The design is claimed to provide relatively high rates of deposition, with high cathode efficiencies.

A plant size evaluation of the system was conducted in 1985-1986 at a zinc electroplating operation (Torcad Ltd. Toronto, Ont.)[24]. The system was installed to recover zinc metal from rinse waters following acid pickling. The waters were treated at a rate of 140 L/hr, and zinc concentration was reduced from 7 g/L to 1 g/L.

The Chrome-Napper system manufactured by Innova Technology Inc. makes use of both low rate evaporation and "Thick membrane" electrolytic ion transfer. Chrome containing rinsewater is circulated through electrolytic cells containing an ion transfer membrane. A continuous direct electric current across the cell electrodes causes chrome ions to be concentrated in an anolyte solution, thus reducing the metal concentration in the rinse water. After reaching a pre-determined maximum concentration, the analyte solution is returned to the platings baths.

A demonstration of the Chrome-Napper system was conducted in 1982 at a nickel-chrome electroplating shop in Toronto (Dovercourt Electro-plating Co. Ltd).[25]. The system resulted in a 50% recovery of chrome which would otherwise be sent to sewer. Final rinse tank chrome concentrations were below 70 mg/L. Anolyte levels of 30,000 mg/L, were achieved thus providing a 750 fold chrome concentration increase.

Celec Inc.[26] conducted laboratory scale tests using synthetic solutions cuprous cyanide (CuCN) and sodium cyanide using a porous carbon electrode. They claimed a 90% cyanide and copper recovery from the solutions, and concluded the electrolytic system to be technically and economically viable.

4.3 SOLVENT EXTRACTION PROCESSES

Solvent extraction (SX) technology is coming into wide use in the mineral processing industry. Two recent additions involve solvent extraction of copper from heap leaching solutions at Gibraltar mine in B.C.[27] and at Phelps Dodge Morenci Inc. in Arizona [28]. At both locations, copper is recovered by electrowinning after SX extraction.

Metal separations by solvent extraction involve a variety of chemical products used to form chelates and organometallic compounds to provide a change from predominantly aqueous to predominantly organic phase solulibility. The SX compounds are usually reactive to some degree with all metals of interest (eg.Cu,Ni,Fe,Zn), with the reactivity dependent in large measure on the prevailing pH. Thus, selectivity in separation efficiency is usually achieved by controlling the pH at which the extraction is conducted.

The SX process involves steps as follows:

- Extraction of metal from the aqueous phase to the organic phase (containing generally 10 to 15% SX reagent dissolved in organic diluent). Multiple stage countercurrent extraction is often used to promote greater process efficiency and to overcome limited metal solubilities in the organic phase.

- Stripping of metal from the organic phase generally using an acid. The barren organic phase is recycled to the extraction step.

The use of solvent extraction for the recovery of metals from metal finishing wastes has been studied by Brooks [29]. Highly efficient metal removal (99.8%) was attained for the extraction of Fe, Cu, Ni, Co, Zn, and Cr from a multimetal solution. Acid stripping with 20% H_2SO_4 provided efficient metal recovery (approximately 99%) only for copper.

Twidwell[30] conducted a two year study investigating the potential of metal value recovery from metal hydroxide sludges from surface finishing effluent treatment. Based on laboratory and pilot scale studies, flowsheets were designed involving the following steps:

- Sulphuric acid leaching
- Iron removal by jarosite precipitation
- Copper recovery by electrowinning or sulphate crystallization
- Zinc removal by solvent extraction, and zinc recovery by crystallization
- Chromium oxidation and recovery by lead chromate precipitation
- Nickel removal by sulphide precipitation or nickel sulphate crystallization
- Final polishing by ion exchange.

He concluded that the application of the above unit operations to multi-component mixed metal sludges has demonstrated that treatment of such material is feasible. Cost analyses indicated two of the flowsheets provided an income sufficient to offset the cost of the treatment process.

5.0 RECYCLE TO MINERAL PLANTS

5.1 CURRENT PRACTICE

A number of Canadian Mineral producers presently recycle effluent treatment sludges in their operations:

At Brunswick Mining and Smelting, Smelting Division (Belledune, N.B.)[31], clarifier underflow from the lime treatment of various smelter effluents is combined with scrubber and process slurry recycled streams, and sent to the smelter sinter plant.

Hudson Bay Mining and Smelting (HBM&S) treats plant process waste streams using lime. The sludge at 10% is blended with concentrates that are sent to the zinc plant. HBM&S estimates that sludge to the process accounts for less then 1% of the zinc produced[32].

Cominco Ltd. operates a treatment system for removal of heavy metals from effluents at the Trail, B.C. smelter. Effluent treatment sludges are recycled to the smelter for metal recovery[14].

5.2 RECYCLE TO SMELTER/REFINERY CONSIDERATIONS

The base metal smelter/refineries operated by the Canadian Mineral Industry are shown in Table 2. The data indicates a total of 1.4 Million tonnes/year copper, 0.56 Million tonnes zinc, and 0.2 Million tonnes lead, produced per year.

Estimates by GML[3] have indicated annual sludge production of 140,000 dry tonnes, mainly produced by the base metal industry. The results of the sludge analyses of 14 industry representative samples

TABLE 2

CANADIAN BASE METAL SMELTERS AND REFINERIES

Metal Type of Facility	Company Location	Annual Production
Copper Smelter	Hudson Bay Mining and Smelting Flin Flon, Man.	68,600 t/y Cu anodes
Copper/Nickel Smelter	Falconbridge Ltd., Falconbridge, Ont.	110,000 t/y Cu-Ni matte
Copper Smelter, Copper/Zinc Refinery	Kidd Creek Mines Ltd Timmins, Ont.	60,000 t/y Cu 65,000 t/y Zn
Copper Smelter	nes Gaspé Murdochville	80,000 t/y Cu anodes
Copper Refinery	Noranda Inc. Noranda, Que	230,000 t/y Cu anodes
Copper Refinery	Canadian Copper Refineries, Montreal, Que	410,000 t/y Cu anodes
Zinc Plant	Inco Ltd Copper Cliff, Ont.	200,000 t/y Cu
Zinc Plant	Cominco Ltd Trail, B.C.	242,100 t/y Zn
Zinc Plant	Hudson Bay Mining and Smelting, Flin Flon, Man.	92,00 t/y Zn
Lead Refinery	Canadian Electrolytic Zinc, Valleyfield, Que	255,000 t/y Zn
Lead Refinery	Cominco, Ltd Trail, B.C	123,300 t/y Pb
Nickel Refinery	Brunswick Mining and Smelting Belledune, B.C.	75,000 t/y Pb
Nickel Refinery	Sherritt Gordon Mines Ltd Fort Saskatchewan, AB.	25,000 t/y Ni
Nickel Refinery	Inco Ltd. Thompson, Man.	
Nickel Refinery	Inco Ltd Copper Cliff, Ont	62,500 t/y Ni
	Inco Ltd Port Colborne, Ont.	100,000 t/y Ni

are presented in Table 3[3]. As these results show, copper nickel and lead were each found to exceed 0.5% of dry solids in only 2 of the samples tested. Zinc was present at concentrations above 0.5% in 10 of the 11 samples, with levels as high as 27.4%. Thus, these results suggest that in most cases, recycle of sludges to a zinc plant would be most beneficial.

Discussions with Noranda Technology Centre personnel[33] regarding technical aspects of sludge recycle to zinc plants have indicated the following:

1. Zinc plant operation is sensitive to impurities in the concentrate (eg. cobalt, nickel, selenium and tellurium). These impurities interfere with the electrolysis process, and require bleed to control excessive levels.

2. Manganese and magnesium are inert components which also have to be controlled in the electrolyte by bleeding.

3. Almost all the sludges were found to contain high iron concentrations, with levels as high as 29.4% Fe. Iron in zinc plant is precipitated (eg. jarosite), thus creating a disposal problem. For example, at Canadian Electrolytic Zinc (CEZ) in Valleyfield, 1100 t/day of zinc calcine produces 110 to 120 t/day Fe and 350 to 400 t/day of residues with 30 to 35% Fe content.

4. The handling of sludges at a zinc plant may create problems. Concentrates received generally contain 6 to 8% moisture. Thus, sludges would require drying to less than 10% moisture for handling, and to minimize effects on the zinc plant water balance.

TABLE 3 - ICP SLUDGE ANALYSIS CONCENTRATIONS GREATER THAN 0.5% wt/wt dry sludge (wt %)

		Al	As	B	Ca	Cd	Cu	Fe	Mg	Mn	Na	Ni	Pb	Sb	Zn
BASE METAL															
BM-1		1.9	-	6.5	1.8	-	-	3.7	2.5	2.1	-	-	-	-	6.3
BM-2		0.63	1.2	2.2	1.1	2.4	-	1.2	5.6	0.7	1.8	-	2.0	-	7.0
BM-3		-	-	1.1	0.8	-	-	-	-	-	-	-	-	-	27.4
BM-4		-	-	-	34.4	-	-	0.6	1.2	-	-	-	-	-	5.2
BM-5		-	-	0.63	16.0	-	-	0.7	3.1	-	3.3	-	-	-	1.4
BM-6		-	-	-	8.2	-	-	29.4	2.1	-	-	-	-	-	2.6
BM-7		-	-	6.6	16.5	-	-	10.1	4.5	0.5	-	6.3	-	-	-
BM-8		-	-	-	14.7	-	-	6.6	4.4	-	-	-	-	-	1.6
BM-9	Fresh	1.0	-	-	2.5	-	-	25.3	7.3	1.5	-	-	-	-	2.4
BM-9	Aged	1.9	-	-	8.1	-	2.5	25.0	3.1	1.8	-	-	0.66	-	2.5
BM-10	Fresh	3.0	-	-	3.9	-	-	19.3	5.1	0.5	-	-	-	-	1.4
BM-10	Aged	2.0	-	-	5.8	-	-	20.5	4.2	-	-	-	-	-	0.9
BM-11		1.2	-	-	8.2	-	-	7.0	1.0	-	-	-	-	-	14.3
PRECIOUS METAL															
PM-1		2.4	-	2.1	20.8	-	-	2.7	1.6	-	-	-	-	-	-
PM-2		7.3	-	-	26.5	-	6.5	38.4	1.2	-	-	1.7	-	5.4	-
PM-3		-	1.5	-	13.6	-	-	16.9	-	-	6.5	-	-	-	-

5.3 RECYCLE TO CONCENTRATORS

Most of the sludge volumes are currently produced by mining/milling operations. Thus, ideally, these sludges would be recycled to the mill. The only Canadian operation which practised some degree of sludge recycle was Mattabi Mine, which blended sludges from its effluent treatment plant into the concentrate thickener. The small sludge volumes produced had little effect on the zinc concentrate quality, which was shipped to CEZ in Valleyfield. Thus, in some circumstances, this method of sludge disposal may be feasable for other operations.

Direct recycle of metal hydroxide sludges to milling, however, does not appear to be feasible. Discussions with personnel of the Noranda Technology Centre[34] regarding technical aspects of internal sludge recycle in mills have indicated the following:

1. Most mill circuits operated by flotation of metal sulphides. Oxidized material will interfere with the flotation process. Thus, direct recycle of metal hydroxide sludges would reduce flotation efficiencies, and metal recoveries.

2. Metal hydroxides produced during lime precipitation form fine particles which may not be flotable in sulphide systems, and would probably report to disposal as tailings. In high sulphide tailings, acid generation occurs frequently, and would result in subsequent disssolution of the metal hydroxides.

Use of sulphide in place of hydroxide precipitation would produce metal sulphides which may be recycled more successfully to flotation.

6.0 PROCESS FLOW SHEETS AND COST ESTIMATES

Two cases representative of Canadian base metal producers were utilized for the evaluations of sludge minimization/utilization flowsheets and costs. As shown in Table 4, Case A involves a highly contaminated effluent, with relatively high metal (1500 mg/L) and sulphate (5000 mg/L) levels. Case B considers an effluent stream containing total metals of 275 mg/L and lower sulphate levels (1200 mg/L). Both effluents were taken to have a flow of 200 m^3/h.

6.1 BASIS FOR CAPITAL COSTS

Capital cost estimates were derived by assuming the facility to be a battery limits add-on installation (ie. construction of a facility on an already prepared and serviced site) to handle the effluents or sludge of an existing mineral producer. The costs considered and excluded for the purpose of this study are summarized in Table 5. It is assumed that the existing utility lines (electricity, power, steam), and all service facilities would be adequate to handle the increased load.

Cost estimates of equipment were derived using two methods. In most cases, individual equipment was sized and costed using either supplier's quotations or recent literature, and the sixth tenths factor rule to generate costs at other than given equipment ratings. In other cases, correlations relating capital cost to process throughput were utilized. All equipment costs are based on 3rd quarter 1988 values, with the Marshall and Swift index used where necessary for updating earlier costs.

TABLE 4

BASE CASE EFFLUENT PARAMETERS

EFFLUENT PARAMETER	CASE A	CASE B
pH	3	2.8
Fe (mg/L)	1000	250
Cu (mg/L)	20	5
Zn (mg/L)	500	20
SO_4 (mg/L)	5000	1200
Flow (m^3/hr)	200	200

The format used for deriving overall capital costs is indicated in Table 6. The bare equipment is the sum of all process units. A contingency of 15% was applied for miscellaneous equipment and the total factored by 1.45 to obtain the installed cost. Other process units where the derived costs includes installation charges were then added. Subsequently, factors for services and building costs were considered. Finally, an allowance for engineering procurement and construction supervision expenses, and contingency costs were used to generate the overall total capital cost.

6.2 BASIS FOR OPERATING AND MAINTENANCE COSTS

The following parameters contributing to operating and maintenance costs were included:

- Reagents (lime, H_2SO_4, etc)
- Utilities (electricity, steam, fuel)
- Operating man-power
- Replacement of components

For each alternative, reagent and utility requirements were based on process conditions with third quarter 1988 unit costs used.

Cost for maintenance, taxes, and insurance were taken as 4.5% of the overall total capital costs. Credits for recovered metals were taken against operating costs.

6.3 ANNUALIZED CAPITAL CHARGES

In order to compare alternative sludge disposal/recovery methods, it is desirable to calculate an annualized capital charge, to be added to the annual operating cost to provide an overall "figure of merit".

TABLE 5

SUMMARY OF BATTERY LIMITS COST COMPONENTS

INCLUDES THE COST OF	EXCLUDES THE COST OF
. all process equipment.	. land, site preparation, fencing, access roads and paving, loading and unloading facilities.
. all material and labour to connect and make operable this process equipment, including control rooms, buildings within the battery limits.	. off-site utilities and piping to/from battery limits from such services.
. indirect field expense, engineering, construction supervision fees.	. Support facilities such as the administration building, laboratory, cafeteria, gate house, maintenances shops, fire fighting equipment, communications systems and safety equipment.

TABLE 6

FORMAT FOR CAPITAL COST CALCULATIONS

NO.	ITEM	$'000
1.	Equipment Items	239
2.	Misc. Unspecified Items (15% of Item 1)	35
	Subtotal Equipment	274
3.	Installed Cost (Equipment x 1.45)	398
4.	Services: (40% of installed cost includes: Electric 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	159
5.	Building	210
	Subtotal Installations	767
6.	Engineering and Construction (25% of Installations	191
	Subtotal	958
7.	Contingency (40% of Item 6)	383
8.	Overall Total Capital Cost	1,343

Annualized capital charges are based on the concept of opportunity cost, i.e. the earnings required to achieve a company's minimum hurdle rate on that capital. The minimum hurdle rate, for risk-free investment, is a function of the company's average cost of capital and is estimated to be 15%, expressed as after-tax discounted cash flow (DCF) return on investment. The effect of taxation must be taken into account, since a capital expenditure provides a "tax shield" spread over a number of future years, whereas an annual cost provides an immediate tax shield. Detailed tax calculations were carried out for hypothetical but realistic cases, from which it is estimated that the net present value of the tax shield is typically 40% of the capital investment, taking into account capital cost allowance, earned depletion and investment tax credit. The annual after-tax earnings required to provide a 15% DCF return on a $100 capital investment is therefore calculated as follows:

$$(100-40)\ \frac{i\ (1+i)^n}{(1+i)^n - 1} = \$\ 9.59$$

Where i, discount rate = 15%
n, project life = 20 years

The pretax earnings, based on the effective tax rate of 52% would be:

$$\frac{\$\ 9.59}{(1-.52)} = \$\ 19.98,$$ which is rounded to 20% of capital

Thus, annualized capital charges are taken as 20% of capital cost.

6.4 METAL PRECIPITATION AND DEWATERING

The use of lime, sodium hydroxide, and sodium sulphide was considered for the treatment of the two base case effluents. Estimates of annual waste production for clarifier underflow, as well as subsequent vacuum filtration and drying were derived using a flowsheet as presented in Figure 4, and are shown in Table 7. The sludge produced would be hauled to an on-site maximum security landfill site, consisting of a double membrane bottom, leachate collection and treatment and leak detection systems, as shown in Figure 5. Filled cells would be sealed with a top membrane layer, a clay cap, and seeded.

As indicated in Table 7, use of lime for the treatment of the high sulphate effluent (Case A) results in over three times the weight of dryer discharge compared to NaOH or Na_2S

Capital and operating cost estimates were prepared for metal precipitation facilities, followed by two levels of dewatering: using vacuum filters only, and vacuum filters and dryers. Estimates for sludge disposal costs were based on the use of a maximum secure landfill, at $75/tonne as operating costs. As shown in Table 8, due to lower lime costs, in all cases considered, lime appears to provide the most cost-effective approach.

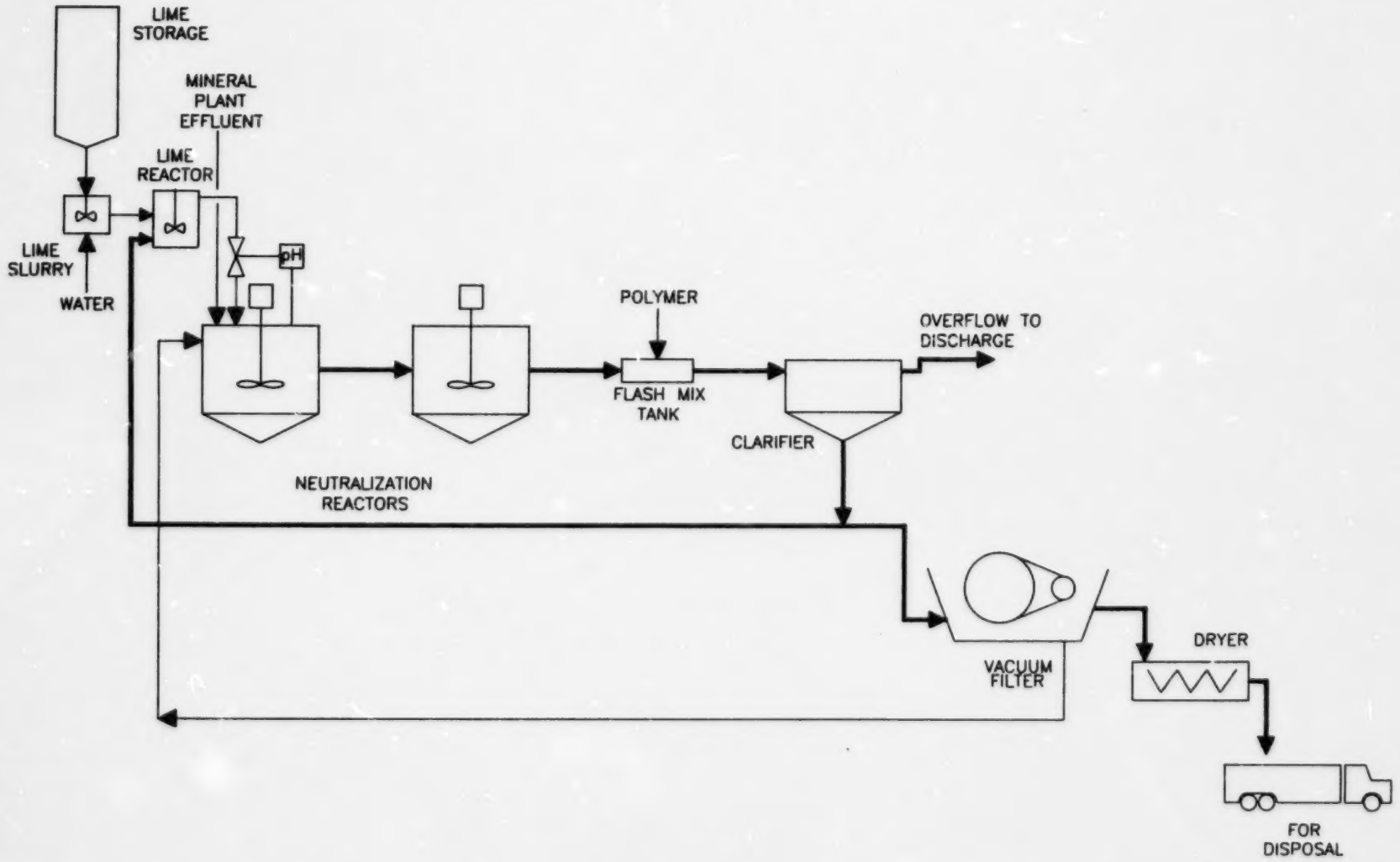

FIGURE 4: EFFLUENT TREATMENT / SLUDGE DEWATERING FLOWSHEET

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TABLE 7

ANNUAL WASTE TONNAGES AS A FUNCTION OF REAGENTS AND DEWATERING

	CASE A			CASE B		
	Clarifier Underflow t/yr	Filter Cake t/yr	Dryer Discharge t/yr	Clarifier Underflow t/yr	Filter Cake t/yr	Dryer Discharge t/yr
Hydrated Lime	81,250	36,100	18,000	4,200	1,900	940
Sodium Hydroxide	95,600	23,900	5,300	16,940	4,240	940
Sodium Sulphide	93,600	23,400	5,200	16,600	4,150	920

BASIS:

Clarifier underflow - 20% solids for lime, 5% solids for NaOH, Na_2S

Filter cake - 45% solids for lime, 20% solids for NaOH, Na_2S

Dryer discharge - 90% solids for lime, NaOH, Na_2S

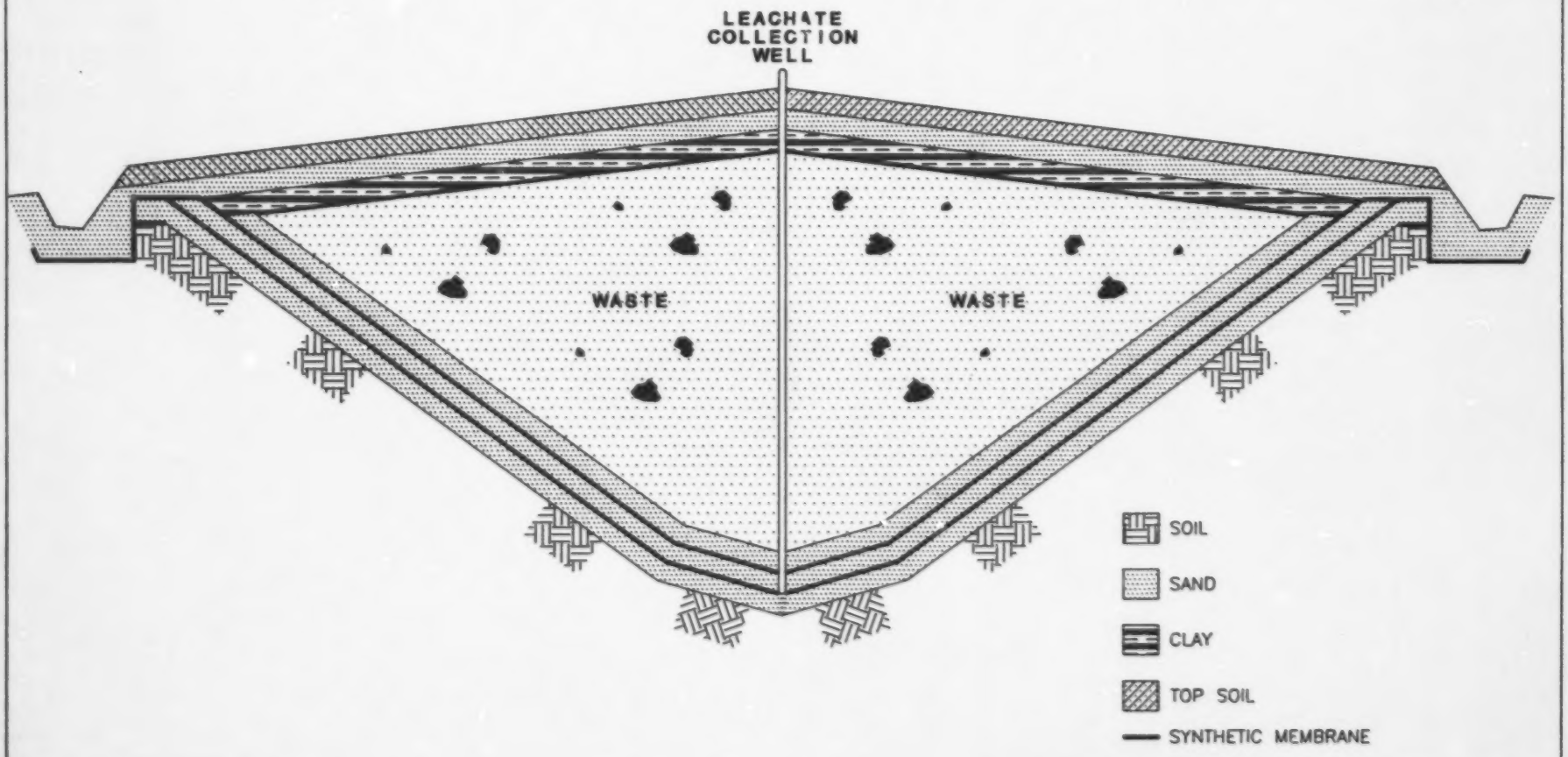

FIGURE 5: MAXIMUM SECURE LANDFILL SCHEMATIC

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TABLE 8

SUMMARY OF COSTS - METAL PRECIPITATION AND SLUDGE DEWATERING

Treatment	Case A					Case B				
	Capital ($Million)	Operating and Maintenance per year ($Million)	Reagents per year ($Million)	Sludge Disposal per year ($Million)	Overall Annual (1) ($Million)	Capital ($Million)	Operating and Maintenance per year ($Million)	Reagents per year ($Million)	Sludge Disposal ($Million)	Overall Annual (1) ($Million)
Precipitation and Filter										
Hydrated Lime	4.0	0.46	0.8	2.7	4.8	2.9	0.34	0.1	0.02	1.0
Sodium Hydroxide	3.1	0.37	2.1	1.8	4.9	2.9	0.34	0.5	0.05	1.5
Sodium Sulphide	3.1	0.37	4.5	1.8	7.3	2.9	0.34	1.2	0.10	2.2
Precipitation Filter and Dryer										
Hydrated Lime	6.8	1.1	0.8	1.4	4.7	3.4	0.4	0.1	0.03	1.2
Sodium Hydroxide	5.9	1.0	2.1	0.4	4.7	4.2	0.5	0.5	0.08	1.9
Sodium Sulphide	5.9	1.0	4.5	0.4	7.1	4.2	0.5	1.2	0.13	2.7

(1) Sum of operating and maintenance, reagent, sludge disposal, and annualized capital.

6.5 ION EXCHANGE/ELECTROWINNING

6.5.1 Process Description

As shown in Figure 6, the ion exchange/electrowinning, sequence is as follows:

The effluent stream is passed through fixed ion exchange columns containing chelating resins. Heavy metals are removed to below 1 mg/L, while iron removal efficiency is 20%. The discharge stream containing dissolved calcium, sulphate and iron is sent out to the receiving stream or sewer (in some cases, additional treatment for iron may be required). The resin is regenerated using a 10% H_2SO_4 solution, which removes 90% of the loaded zinc, copper and iron.

The resin is subsequently washed with 9 to 10 volumes water to remove the balance of the metals. The rinse water is recycled to the column feed. The sulphuric acid wash solution is sent to pH adjustment with NaOH, to raise the pH to 2. Subsequently, the solution is passed through a selective ion exchange bed for copper removal. The copper free solution is sent to zinc electrowinning circuit.

The selective ion exchange resin is regenerated using a H_2SO_4 acid solution. The acid wash is sent to copper electrowinning. A column rinse (9 to 10 volumes water) is used to remove the remaining copper, which is recycled to feed the chelating resin beds. The copper and zinc electrowinning circuits result in leaving an effluent containing 250 to 300 mg/L metal. The copper circuit overflow is recycled to feed of the chelating resin beds. The zinc circuit overflow contains iron which is

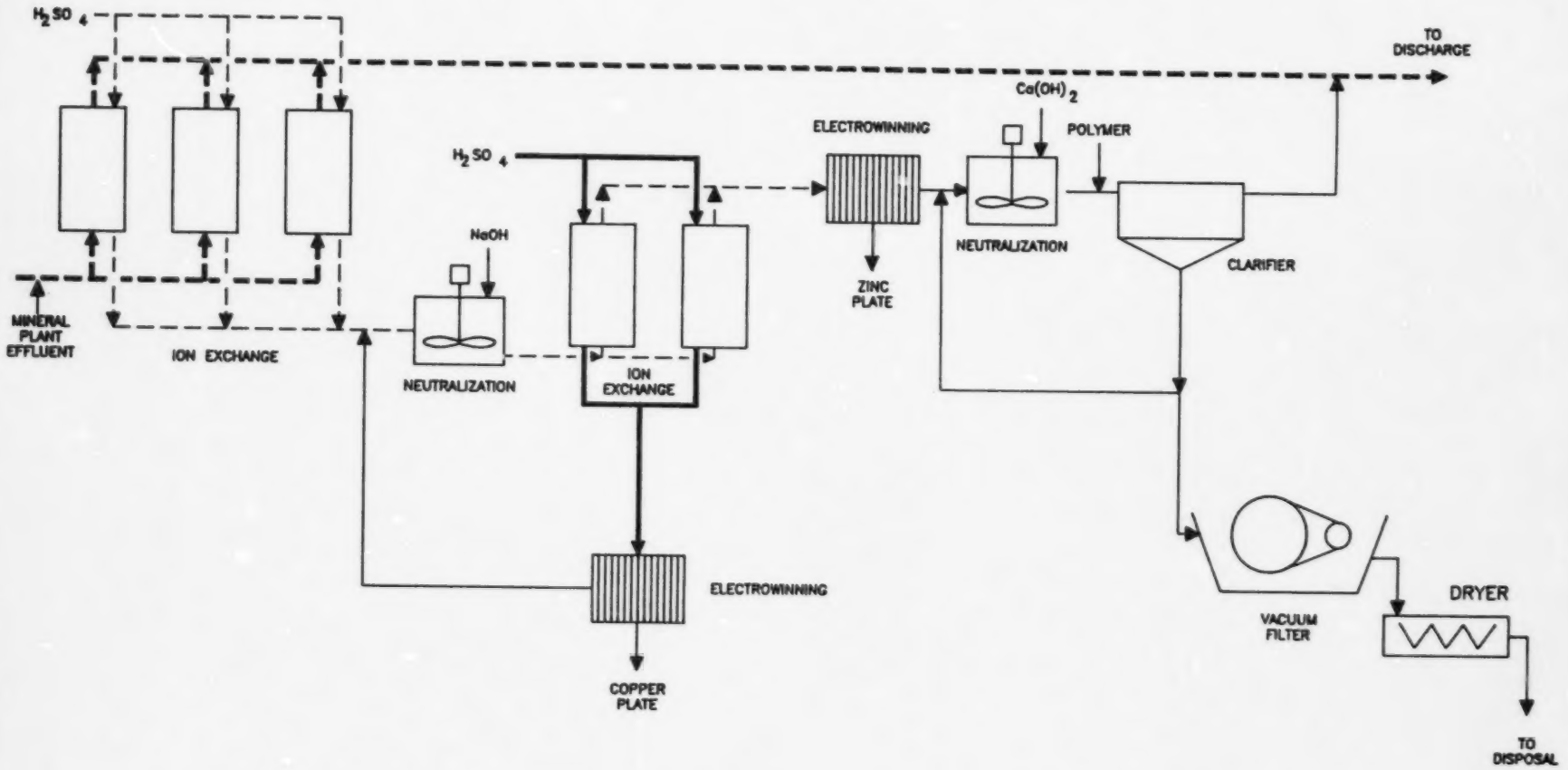

FIGURE 6: ION EXCHANGE / ELECTROWINNING FLOWSHEET

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not removed in the circuits. This flow will be discharged for treatment with lime to precipitate the zinc and iron , and neutralize the sulphuric acid. The precipitated sludge can be filtered, dried, and disposed. Because of the presence of zinc, this sludge would be classified as a hazardous waste in many juridictions, and would require secure landfill.

6.5.2 Cost Estimates

The estimates of capital and operating costs are summarized in Table 9. As shown, capital costs are $11.2 Million and $2.4 Million for case A and B installations respectively. Corresponding overall annual costs (capitalization, operating, and maintenance) are $4.5 Million and $1.7 Million, assuming a credit for recovered copper and zinc, of $1.4 Million and $0.08 Million respectively.

6.6 SOLVENT EXTRACTION/ELECTROWINNING

6.6.1 Process Description

The process sequence is illustrated in Figure 7. Metal hydroxide sludges are initially leached with sulphuric acid to dissolve the metals. Subsequently, the leach liquor is reacted with sodium carbonate at a ratio of 4:1 Na_2CO_3: Fe to precipitate iron as a jarosite. The jarosite solids are separated from the leach solution in a clarifier, then dewatered to roughly 10% moisture using a filter and a dryer. (It is expected that the jarosite will contain coprecipitated zinc and copper, and will require disposal to an approved site).

TABLE 9

SUMMARY OF COSTS - ION EXCHANGE/ELECTROWINNING

PARAMETER	Case A ($ Million)	Case B ($ Million)
Capital	11.2	2.4
Annual		
Operating and Maintenance	2.7	0.4
Annualized Capital	2.2	0.5
Sludge Disposal (1)	0.2	0.04
Lime Neutralization	0.8	0.8
Credit for Recovered Metals		
Copper	0.12	0.03
Zinc	1.3	0.05
	(1.42)	(0.08)
Overall Annual	4.5	1.7

(1) To maximum secure landfill, at $ 75/tonne

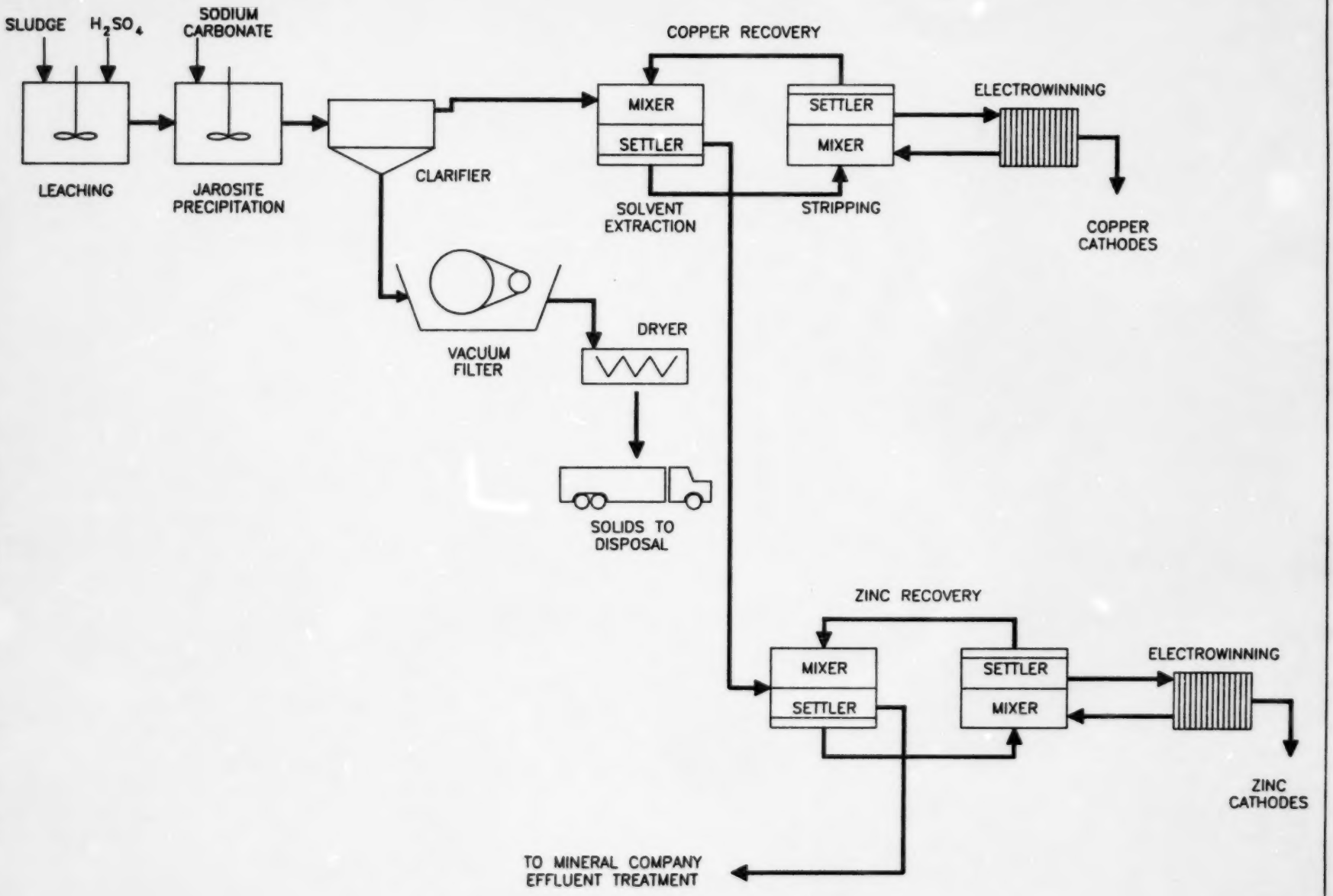

FIGURE 7: SOLVENT EXTRACTION / ELECTROWINNING FLOWSHEET

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The leach solutions are sent to separate solvent extraction circuits for zinc and copper recovery. In each case, the organic phase is stripped with sulphuric acid, and the barren solution is recycled to the first stage. The barren acidic leach liquor following copper and zinc removal is sent to the plant's effluent treatment system for neutralization, and discharge.

The metal containing acid streams from each of the SX circuits are sent for metal recovery by electrowinning.

6.6.2 Cost Estimates

The cost estimates of solvent extraction/electrowinning are presented in Table 9. Capital costs are $12.8 Million and $2.5 Million respectively for case A and case B facilities. Overall annual costs (including capitallization, operating and maintenance) when credit for recovered metals is allowed, are $8.4 Million and $2.2 Million/year. The case A costs assume the use of sodium hydroxide in the metal precipitation step. The use of calcium hydroxide would result in total gypsum and jarosite solids quantities exceeding those produced by lime precipitation alone.

TABLE 10

SUMMARY OF COSTS - SOLVENT EXTRACTION/ELECTROWINNING

PARAMETER	Case A ($ Million)	Case B ($ Million)
Capital	12.8	2.5
Annual		
Operating and Maintenance	2.7	0.5
Annualized Capital	2.6	0.5
Lime Neutralization	4.2	1.2
Sludge Disposal (1)	0.3	0.08
Credit for Recovered Metals		
Copper	0.12	0.03
Zinc	1.3	0.05
	(1.32)	(0.08)
Overall Annual	8.4	2.2

(1) To maximum secure landfill, at $ 75/tonne

7.0 DISCUSSION AND RECOMMENDATIONS

The alternate methods of dealing with sludge generation in the mineral industry will all result in significant costs which are not offset by the value of the metals recovered at current metal prices. A summary of the total annual costs (operating and maintenance, annualized capital, sludge disposal, etc) for the options considered is summarized in Table 11, and indicates the following:

- The alternatives to conventional lime treatment will all result in significant cost increases to the mineral industry.

- Both capital and operating costs are significantly affected by contaminant concentrations.

- Ion exchange costs are lower in some cases than combined conventional lime precipitation, sludge dewatering, and disposal.

The methods of dealing with sludge generation will all result in residues for ultimate disposal. A maximum secure landfill has recently been accepted by the Quebec Ministère de l'Environnement as a viable disposal for an industrial waste containing leachable heavy metals [35]. The system involves a double membrane at the bottom of the disposal cell, with a leachate collection/treatment system. Such a system may be acceptable for mineral industry sludges.

Because ion exchange treatment appears a cost-effective option for the treatment of mineral industry effluents, it is recommended that the following be addressed:

- Technical evaluation of currently available selective resins for the treatment of mineral industry effluents with respect to such operating parameters as metal loadings, effects of contaminants on resin life and performance.

- The development of cheaper and more specific, ion exchange resins.

The possibility and overall technical and costs aspects of the recycle of sludges to smelters and refineries need further elucidation. Although, as indicated above, some recycle of sludges to smelter is practised in Canada, little information exists on the amounts of sludges recycled, and the pathways of the various sludge components through the smelter.

The overall technical and costs aspects of recycle of sludges to zinc refineries also need addressing. Disposal to zinc plants appears to be beneficial because of the high zinc content in the sludges. However, iron in the sludges would report to jarosite. Because of carry-over of heavy metals with jarosite, this creates a hazardous waste disposal problem at the zinc plant. Also, potential sludge contaminants which may affect zinc plant operation need to be identified.

TABLE 11

COST SUMMARY, EFFLUENT AND SLUDGE DISPOSAL OPTIONS

Item	Case A Total Annual Cost ($ Million)	Case B Total Annual Cost ($ Million)
1. Lime Precipitation only (1)	1.6	0.9
2. Metal Precipitation (2)		
2.1 Hydrated Lime	4.7	1.2
2.2 Sodium Hydroxide	4.7	1.9
2.3 Sodium Sulphide	7.1	2.7
3. Ion Exchange/Electrowinning (3)	4.5	1.7
4. Solvent Extraction/Electrowinning (3)	8.4	2.2

(1) Not cost for disposal of sludges, reflects current practices.
(2) Precipitation, filter and dryer dewatering, and disposal in a maximum security landfill.
(3) Includes disposal of residues and sludges in a maximum security landfill.

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33. Personal Communication, Noranda Technology Centre, P. Spira, January 28, 1988.

34. Personal Communication, Noranda Technology Centre, R. Stanley, Feb. 5, 1988.

35. Gionet Mellor Liebich Associates Limited, Internal Projects, Nov. 1988.

APPENDIX 1

RESULTS OF LEACHING STUDIES [3]

LEACHING TEST RESULTS

The results of the leaching studies of mineral industry wastewater treatment sludges are presented in table 11 and indicate the following:

- The release of contaminants from the sludges is highly related to the final pH of the leachates.

- When distilled water is used as the leaching medium, final pH values were generally alkaline and above pH 8, indicating substantial residual alkalinity in the sludges.

- The use of acetic acid/sodium acetate solution does not appear to result in higher leach results than sulphuric acid. Final leach pH values are indicated as the predominant factor of contaminant levels.

- Contaminant release in leach liquor was also related to initial concentration of metals in the sludge. Thus, lead, cadmium, nickel, and arsenic were present in low concentrations in most sludge samples, as well as leach solutions. Where these elements are present at levels above 0.5% (e.g. BM-2, PM-2 and BM-7), low leach pH solutions contained significantly higher contaminant concentrations.

- The highest release and leach solutions concentrations were for zinc, iron, and copper, with levels as high as 5200 mg/L Zn, 770 mg/L Fe, and 93 mg/L Cu.

- Both old and fresh sludges appeared to result in similar leach contaminant concentrations, (i.e. BM-9, BM-10).

TABLE 11 – LEACHATE ANALYSIS

	LEACHING PROCEDURE *	Initial pH	Final pH	Pb (mg/L)	Cu (mg/L)	Ni (mg/L)	Zn (mg/L)	Fe (mg/L)	Sb (mg/L)	As (mg/L)	CN (mg/L)	Cd (mg/L)	Cr (mg/L)
BASE METALS													
BH-1	L1	4.00	8.96	0	0	0	0	0	-	-	-	0	-
	L2	2.50	6.70	0	0	0	16.0	0.3	-	-	-	0.1	-
	L3	4.00	4.00	0	0.3	0.2	95.0	0.5	-	-	-	0.1	-
	L4	2.30	2.30	0	0.4	0.3	120.0	45.0	-	-	-	0.1	-
	L5		4.66	0	0.3	0	90.0	2.0	-	-	-	0.1	-
	L6		9.83	0	0	0	0	0	-	-	-	0	-
BH-2	L1	4.00	9.86	0	0.3	0	0	0	-	0.006	-	0	-
	L2	2.50	8.90	0	0.3	0	0	0	-	-	-	0.2	-
	L3	3.90	3.90	6.7	0.3	0.5	220.0	1.0	-	-	-	63.0	-
	L4	2.67	2.67	4.2	0.4	0.6	225.0	4.1	-	2.7	-	63.0	-
	L5		5.47	12.0	0.2	0.5	208.0	0.3	-	0.37	-	59.0	-
	L6		10.39	0	0	0	0	0	-	-	-	0.1	-
BH-3	L1	4.00	8.30	0	0		1.3	0	-	-	-	0.1	-
	L2	2.50	6.96	0	0	0	7.5	0	-	-	-	0.7	-
	L3	4.05	4.05	1.5	21.0	0.3	1900.0	1.0	-	-	-	3.2	-
	L4	2.56	2.56	2.0	21.0	0.3	1900.0	6.4	-	-	-	3.2	-
	L5		5.52	2.0	18.0	0.4	1900.0	0.3	-	-	-	2.7	-
	L6		8.77	0	0	0.2	2.7	0	-	-	-	<0.1	-
BH-4	L1	4.00	9.39	0	0	0	0	0				0	0
	L2	2.50	8.72	0	0.1	0	0	0				0	0
	L3	4.10	4.10	0	12.0	0.8	970.0	2.5	-	-	-	2.1	0
	L4	2.59	2.59	0.5	13.0	1.0	970.0	22.0	-	-	-	2.2	0
	L5	4.50	6.25	0	2.0	0.7	680.0	0	-	-	-	0.4	0
	L6		9.85	0	0	0	1.5	0	-	-	-	0	0
									-	-	-	-	-
BH-5	L1	4.00	9.32	0	0	0	0	0	-	-	-	0	-
	L2	2.50	8.87	0	0	0	0	0	-	-	-	0	
	L3	4.00	4.21	0	2.0	0.8	310.0	3.0	-	-	-	0.4	-
	L4	2.50	2.58	0.5	3.0	0.9	375.0	25.0	-	-	-	0.4	-
	L5	4.50	5.30	0	1.0	0.7	400.0	1.0	-	-	-	0.4	-
	L6		8.47	0	0	0	0.8	0	-	-	-	0	-
BH-6	L1	4.00	8.20	0	0	0	0	0	-	-	-	0	-
	L2	2.50	5.42	0	0.7	0	13.0	2.0	-	-	-	0.1	-
	L3	4.00	4.13	-	0.3	0	34.0	2.5	-	-	-	0.1	-
	L4	2.50	2.40	0	0.3	0.3	30.0	87.0	-	-	-	0.1	-
	L5	4.50	4.60	0	0.3	0.2	21.0	1.0	-	-	-	0.1	-
	L6		7.88	0	0	0	0	0	-	-	-	0	-
BH-7	L1	4.00	8.66	0	0	1.2	0	0.5	-	-	-	0	-
	L2	2.50	6.78	0	0	9.5	0	0.5	-	-	-	0	-
	L3	4.00	4.24	0	0.8	28.3	0.6	0.3	-	-	-	0	-
	L4	2.50	2.40	0	1.1	33.0	0.6	42.0	-	-	-	0	-
	L5	4.50	4.64	0	1.0	27.5	1.1	1.0	-	-	-	0	-
	L6		8.91	0	0	1.0	0	0.5	-	-	-	0	-
BH-8	L1	4.00	8.98	0	0.3	0	0.9	0	-	-	-	0	-
	L2	2.50	3.16	0	0.7	0	5.9	6.1	-	-	-	0	-
	L3	4.00	4.03	0	0.3	0	5.7	3.1	-	-	-	0	-
	L4	2.50	2.67	0.1	0.5	0.3	6.1	22.0	-	-	-	0	-
	L5	4.50	4.61	0.2	0.5	0.2	6.5	2.0	-	-	-	0	-
	L6		9.05	0	0	0	0.1	0	-	-	-	0	-
BH-9	L1	4.00	9.10	0	0	0	0.1	0	-	-	-	-	0
	L2	2.50	8.87	0	0	0	0.1	0	-	-	-	-	0
	L3	4.00	4.34	1.0	11.0	2.3	365.0	0	-	-	-	1.5	-
	L4	2.50	2.72	3.0	66.0	2.5	420.0	125.0	-	-	-	1.8	-
	L5	4.50	7.87	1.0	0	0	0.1	0	-	-	-	0	-
	L6		9.23	0	0	0	0.1	0	-	-	-	0	-
BH-9-0	L1	4.00	8.27	0	0	0	0.1	0	-	-	-	0	-
	L2	2.50	8.17	0.3	0	0	0.1	0	-	-	-	0	-
	L3	4.00	4.27	1.0	0.8	4.5	850.0	0.5	-	-	-	1.2	-
	L4	2.50	3.29	4.0	7.5	5.5	1265.0	6.0	-	-	-	2.3	-
	L5	4.50	7.40	0.7	9	9	0.2	0	-	-	-	0	-
	L6		9.17	0	0	0	0.1	0	-	-	-	0	-
BH-10	L1	4.00	9.28	0	0	0	0	0.5	-	-	-	0	-
	L2	2.50	9.10	0	0	0	0.1	0.5	-	-	-	0	-
	L3	4.00	4.10	0.3	18.6	1.0	232.0	1.0	-	-	-	0.4	-
	L4	2.50	2.54	0.8	49.0	1.3	266.0	770.0	-	-	-	0.5	-
	L5	4.50	7.17	0.4	0	0.2	1.4	0.5	-	-	-	0	-
	L6		9.17	0	0	0	0	0	-	-	-	0	-

* Leaching Procedure – see note end of Table

TABLE 11 cont'd - LEACHATE ANALYSIS

		Initial pH	Final pH	Pb (mg/L)	Cu (mg/L)	Ni (mg/L)	Zn (mg/L)	Fe (mg/)	Sb (mg/L)	As (mg/L)	CN (mg/L)	Cd (mg/L)	Cr (mg/L)
BH-10-0	L1	4.00	9.14	0	0	0	0	0.5	-	-	-	0	-
	L2	2.50	8.88	0	0	0	0.1	0.5	-	-	-	0	-
	L3	4.00	4.22	0	6.6	0.7	170.0	2.0	-	-	-	0.3	-
	L4	2.50	2.73	0.4	58.0	0.8	330.0	220.0	-	-	-	0.7	-
	L5	4.50	7.90	0.4	0	0.2	0.1	0	-	-	-	0	-
	L6		9.03	0	0	0	0	0	-	-	-	0	-
BH-11	L1	4.00	11.07	0	0	0	0.3	0	-	-	-	0	-
	L2	2.50	11.07	0	0	0	0.2	0	-	-	-	0	-
	L3	4.00	4.15	0	55.0	1.3	5120.0	1.0	-	-	-	18.0	-
	L4	2.50	2.56	1.9	93.0	1.5	5380.0	690.0	-	-	-	21.0	-
	L5	4.50	10.49	0.4	0	0.3	0.1	0.5	-	-	-	0	-
	L6		11.16	0	0	0	0	0	-	-	-	0	-
PRECIOUS METALS													
PM-1	L1	4.00	10.67	0	0.2	0	0.1	1.5	-	-	-	0	-
	L2	2.50	10.06	0	0.2	0	0	0.5	-	-	-	Tr	-
	L3	4.00	4.21	0	11.0	1.3	7.6	0	-	-	0.02	0.1	-
	L4	2.50	2.62	0	13.0	1.5	8.1	67.0	-	-	-	0.1	-
	L5	4.50	4.72	0	11.0	1.0	7.5	14.0	-	-	0.02	0.1	-
	L6		9.30	0	0	0	0	0	-	-	-	0	-
PM-2	L1	4.00	9.40	0	1.7	0.5	0.1	8.6	1.1	-	-		-
	L2	2.50	7.98	0	0.4	0.3	0	1.5	-	-	-	0	0
	L3	4.00	4.13	0	40.0	18.3	2.0	0.3	-	-	0.8	0	-
	L4	2.50	2.56	0	55.0	21.1	2.4	94.0	0.014	-	-	-	-
	L5	4.50	4.72	0	39.0	13.6	2.3	0.5	0.11	-	0.1	0	0
	L6		7.32	0	0.2	0	0	1.0	-	-	-	0	0
PM-3	L1	4.00	7.30	0	0	0	0	0	-	0.012	-	0	-
	L2	2.50	3.40	0	4.4	0	1.6	1.0	-	-	-	0	-
	L3	4.0	4.02	0	3.4	0	1.5	0	-	-	0.06	0	-
	L4	2.50	2.57	0	4.7	0	1.6	55.0	-	0.118	-	0	-
	L5	4.50	4.57	0.3	3.2	0.2	1.4	0	-	0.014	0.05	0	-
	L6		7.50	0	0	0	0	0	-	-	-	0	-

* LEACHING PROCEDURE

L1: 1/10 dilution of sludge in H_2SO_4 solution at initial pH 4

L2: 1/10 dilution of sludge in H_2SO_4 solution at initial pH 2.5

L3: 1/10 dilution of sludge in H_2SO_4 solution maintained pH 4

L4: 1/10 dilution of sludge in H_2SO_4 solution maintained at pH 2.5.

L5: 1/10 dilution of sludge in acetic acid & sodium acetate solution at initial pH 4.5.

L6: 1/10 dilution of sludge in distilled water.

OTHERS NOTATION

(-) Not determined
(0) Below detection limit